

INTEGRATED NGL RECOVERY AND LNG LIQUEFACTION

This application claims priority to our copending U.S. provisional patent application with the serial number 60/673518, which was filed April 25, 2005.

Field of The Invention

5 The field of the invention is natural gas liquids (NGL) recovery and LNG liquefaction, and particularly integrated plant configurations for same.

Background of The Invention

10 While the crude oil supply in the world is diminishing, the supply of natural gas is still relatively abundant in many parts of the world. Natural gas is typically recovered from oil and gas production wells located onshore and offshore. Most typically, natural gas predominantly comprises C1 (methane). Depending on the particular formations and reservoirs, natural gas also contains relatively low quantities of non-methane hydrocarbons, including C2, C3, C4, C5, and heavier components. Still further components of natural gas include water, nitrogen, carbon dioxide, hydrogen sulfide, mercaptans, and other gases.

15 Natural gas from wellheads is commonly treated and processed, and transported to gas processing plants in high pressure transmission pipelines. However, and especially in remote locations without the necessary pipeline infrastructure, natural gas is commonly transported by liquefying the natural gas and moving the liquefied gas (e.g., using LNG cargo carriers). Unfortunately, direct liquefaction of natural gas is often problematic as natural gas often
20 contains C5, aromatics, and heavier hydrocarbons, which solidify when cooled to cryogenic temperatures. Consequently, such heavier components must be removed to a relatively low level (typically less than 1 ppmv) to avoid solidification and ultimately plugging of the cryogenic heat exchange equipment. Additionally, lighter hydrocarbons such as C2, C3, and C4 must also be removed at least to some degree for the north American market, which
25 typically requires the natural gas to meet a heating value of between 1050 to 1070 Btu/SCF. There are also economic incentives to extract components as they can be sold at a premium price over natural gas. For example, C2 can often be used as a feedstock for petrochemical manufacture, while C3 and C4 can be sold as LPG fuels and C5+ hydrocarbons can be used for gasoline blending.

30 There are numerous configurations and methods known in the art for C2 and C3+ NGL recovery from a natural gas feed. However, past efforts have focused on removal of the

NGL hydrocarbons from natural gas using standalone NGL recovery plants, which operate independently from LNG liquefaction plants. These processes generally produce a relatively low pressure residue gas at ambient temperature, which would necessitate re-compression and re-cooling of the residue gas in an LNG liquefaction plant. Typical examples include various expander processes described in U.S. Pat. Nos. 4,157,904 and 5,275,005 to Campbell et al., 4,251,249 to Gulsby, 4,617,039 to Buck, 4,690,702 to Paradowski et al., 5,799,507 to Wilkinson et al., and 5,890,378 to Rambo et al.

For high C2 recovery, some configurations as described in U.S. Pat. Nos. 6,116,050, require letting down a portion of the residue gas compressor discharge to the NGL recovery column as a methane rich reflux using the Joule-Thomson (JT) valve. While these processes can improve C2 recovery to some extent, additional residue gas compressor horsepower is required, which may render the process costly to operate. There are also more recent advances in the C2 and C3 recovery area (see *e.g.*, commonly owned U.S. Pat. No. 6,837,070) in which a high pressure absorber is coupled with a lower pressure distillation column to improve NGL recovery efficiency. However, these NGL processes are designed for either high C2 or C3 recovery, and generally not designed for varying levels of C2 recovery without lowering C3 recovery. Thus, in most cases, standalone NGL recovery plants are used to produce a low pressure and ambient temperature residue gas that requires re-compression and re-cooling in the LNG liquefaction plant, thereby duplicating many of the refrigeration and heat exchange equipment in the NGL recovery plant.

In other known approaches, attempts were made to include the NGL recovery process as part of the LNG liquefaction plant, as disclosed in the U.S. Pat. Nos. 6,401,486 to Lee et al. and 6,662,589 to Roberts et al. Lee et al. teaches that a methane rich stream can be used as reflux to the NGL recovery column that is coupled with an overhead condenser in another NGL column to achieve an propane recovery of 95%. However this configuration requires the NGL column to operate at a pressure of 450 psig or even lower as the separation of NGL becomes increasingly difficult at the higher pressures due to the correspondingly reduced relative volatility. Consequently, these processes require significant recompression from the column overhead to the required LNG liquefaction pressure, typically from 450 psig to about 800 psig to 900 psig.

In still further known examples, Roberts et al in Pat. No. 6,662,589 teach a C2 rich liquid being recycled from the NGL fractionation unit that is used for C3 absorption in a high pressure absorption column. While this process attempts to operate the NGL column at a high pressure (*e.g.*, 600 psig), NGL separation efficiency suffers as the relative volatility of the NGL components is reduced, which results in recovering significantly less NGL components, especially C2 components. Without removal of a high level of C2 and C3 components, the currently known processes cannot produce a lean natural gas with sufficiently lower heating value content to meet the North American pipeline specifications in an economically manner. Additionally, the lean gas pressure from such known processes would require significant refrigeration in the LNG liquefaction plant due to the relatively low feed gas pressure (LNG liquefaction generally requires significantly less refrigeration duty when operating at a higher pressure, between 800 psig and 900 psig or higher). Additional configurations with similar problems are described in U.S. Pat. No. 5,685,170 to Sorensen and U.S. Pat. App. No. 2005/0247078 to Wilkinson et al.

Thus, while numerous compositions and methods for NGL recovery are known in the art, all or almost all of them, suffer from one or more disadvantages. Therefore, there is still a need for improved NGL recovery, and especially where the NGL plant is integrated with or coupled to a LNG liquefaction unit.

Summary of the Invention

The present invention is directed to configurations and methods of NGL recovery, preferably coupled to an LNG liquefaction process, in which recovery of C2 components can be adjusted using flow ratios of selected process streams. Most preferably, the absorber in such configurations and methods is operated at significantly higher pressure than the distillation column to provide a cryogenic pressurized gas, while the absorber and distillation column temperatures are adjusted to such that desired quantities of C2 and C3+ products are recovered in the NGL. Cryogenic absorber overhead product is then compressed to a pressure suitable for liquefaction using energy derived from expansion of a vapor portion of the feed gas.

In one aspect of the inventive subject matter, a plant includes an absorber configured to receive an absorber feed stream and a first and a second reflux stream, and that is further configured to provide a bottom product stream. A distillation column is configured to receive

a first portion of the bottom product stream and a second portion of the bottom product stream at different points, wherein the distillation column is further configured to operate at a pressure that is lower than an operating pressure in the absorber. Contemplated plants will further include a control unit having one or more control valves that are configured to control flow ratios of (a) the feed stream to the second reflux stream and (b) the first portion of the bottom product stream to the second portion of the bottom product stream, wherein the flow ratios are a function of desired ethane recovery in the distillation column bottom product stream.

It is generally preferred that in such plants a heat exchanger and/or a reflux condenser are configured to heat the first portion of the bottom product stream, and that an expansion device cools the second portion of the bottom product stream. It is also generally preferred that the distillation column is configured to produce a distillation column overhead, and that the plant further comprises a compressor that compresses the distillation column overhead to at least absorber pressure. Where desired, the so compressed overhead is cooled and used as first reflux. Typically, the absorber produces an absorber overhead product that has a temperature of equal or lower than -90 °F and a pressure of between 500 psig and 700 psig, which may be further compressed by a compressor to a pressure suitable for liquefaction of the absorber overhead product (*e.g.*, at least 800 psig). Most preferably, the compressor is operationally coupled to an expander that expands the absorber feed stream.

In another aspect of the inventive subject matter, a method of processing a gas includes a step of providing an absorber that receives an absorber feed stream and a first and a second reflux stream, and that produces a bottom product stream. In another step, the absorber is fluidly coupled to a distillation column such that a first portion of the bottom product stream and a second portion of the bottom product stream are fed to the distillation column at different points, and in another step the distillation column is operated at a pressure that is lower than an operating pressure of the absorber. In yet another step, flow ratios of (a) the feed stream to the second reflux stream and (b) the first portion of the bottom product stream to the second portion of the bottom product stream is controlled as a function of the desired ethane recovery in the distillation column bottom product stream.

In especially preferred methods, the distillation column overhead product is fed to the absorber, most preferably as compressed and cooled first reflux stream. In further preferred

methods, the distillation column is operated at a pressure between 300 psig and 500 psig, and the absorber is operated at a pressure of between 500 psig and 800 psig. Typically, a cooled feed gas is separated into a liquid portion and a vapor portion, and a portion of the liquid portion is after at least partially depressurized and heated and fed into the distillation column.

5 The vapor portion in such methods is preferably split into a first and second stream to thereby form the second reflux stream and the absorber feed stream. It is further preferred that the cryogenic absorber overhead stream is compressed to a pressure suitable for liquefaction using energy from the expansion of the absorber feed stream.

Therefore, and viewed from a different perspective, a method of variably recovering
10 C2 from a feed gas includes a step of feeding an expanded and heated liquid portion of a feed gas to a distillation column and feeding a vapor portion of the feed gas to an absorber. In another step, a flow ratio of an absorber feed to a second reflux to the absorber is controlled, and a first reflux that is provided by a distillation column overhead product is used to thereby control an absorber overhead temperature, and the degree of C2 recovery. In yet another step,
15 the temperature of the absorber bottom product that is fed to the distillation column is adjusted to thereby control distillation column overhead temperature, controlling the desirable NGL recovery levels. The absorber is typically operated at a higher pressure than the distillation column.

Most typically, the step of adjusting the absorber bottom product is performed by
20 heating at least one portion of the absorber bottom product in a heat exchanger, while the step of adjusting the absorber bottom product is performed by cooling at least another portion of the absorber bottom product using a JT valve. In further preferred aspects, the step of adjusting the flow ratio of the absorber feed to the second reflux to the absorber is a function of desired C2 recovery.

25 Various objects, features, aspects and advantages of the present invention will become more apparent from the following detailed description of preferred embodiments of the invention.

Brief Description of the Drawings

Prior Art Figure 1 is a schematic of an exemplary known plant.

Figure 2 is a schematic of an exemplary plant configuration according to the inventive subject matter.

Figure 3 is a graph depicting composite heat curves for heat exchanger 51 and 54 in a plant according to Figure 2 for C3 recovery.

5 Figure 4 is a graph depicting composite heat curves for heat exchanger 51 and 54 in a plant according to Figure 2 for C2 recovery.

Figure 5 is a graph comparing the relative volatilities of the NGL components at 600 psig between the prior art and a plant according to Figure 2 in C2 recovery.

Detailed Description

10 The inventor has discovered that C2 and C3+ components can be effectively and economically separated from natural gas in a plant (which is preferably coupled to an LNG liquefaction plant) using an absorber that operates at high pressure and produces a cryogenic pressurized lean gas, while a distillation column located downstream of an absorber operates at low pressure and produces the NGL as a bottom product and a reflux stream for the
15 absorber. In especially preferred configurations and methods, recovery of C2 in the NGL can be adjusted by controlling process streams within the plant. Moreover, it should be appreciated that the cryogenic absorber overhead product stream is already at relatively high pressure, and energy for recompression to a pressure suitable for liquefaction is typically provided by expansion of a vapor portion of the natural gas feed stream.

20 In contrast, as depicted in **Prior Art Figure 1**, a C2 NGL recovery plant has a single column operating at lower pressure typically 450 psig or lower, which necessitates substantial recompression for delivery of the natural gas to the liquefaction plant. Here, contaminants-free and dried feed gas stream 1, typically supplied at about 1200 psig, is cooled in exchanger 51 using column overhead vapor side reboiler stream 22 and external refrigerant 32. Liquid is
25 removed from separator 52 and sent to the NGL column 58 that acts as a demethanizer. The flashed vapor from the separator 52 is split into two portions, one portion is cooled in the exchanger 54 to provide reflux to the column, while the other portion is expanded in turbo-expander 64, cooled and sent to the lower section for rectification. It should be especially noted that the above gas subcooled process produces a residue gas at ambient temperature and
30 about 550 psig which must be recompressed using re-compressor 100 prior to the standalone

LNG plant. Thus, it should be recognized that such non-integrated NGL recovery plant is inefficient as the residue gas pressure is not adequate to meet the LNG liquefaction plant requirement, requiring additional compression and that the residue gas must be re-chilled in the LNG liquefaction plant.

5 Such difficulties can be overcome in configurations according to the inventive subject matter in which a high pressure cryogenic vapor stream predominantly comprising methane is produced from a NGL recovery plant that advantageously reduces, or even eliminates the duplication of refrigeration and heat exchange steps and equipment in the LNG liquefaction plant. Such configurations and methods greatly reduce the refrigeration requirement in the
10 LNG liquefaction plant, while advantageously allowing recovery of 99% propane and of up to 85% ethane from the feed gas. Contemplated NGL recovery plants will typically produce a lean gas predominantly comprising methane with a predetermined heating value (*e.g.*, to meet the North America natural gas pipeline requirement). Moreover, contemplated plants and configurations can be integrated with an LNG liquefaction plant, which will in turn increase
15 the throughput of the LNG liquefaction train for the same energy input. Still further, it is noted that contemplated configurations can be changed from C3 recovery to C2 recovery by adjusting a flow ratio between the top reflux and the expansion flow while diverting at least a portion of the absorber bottoms product flow to a distillation column.

 In especially preferred configurations and methods, the gas processing portion of an
20 integrated plant comprises a refluxed absorber producing a bottom stream and receiving a feed gas and an absorber reflux stream that is produced from the overhead vapor from a distillation column (preferably after the overhead vapor is compressed and cooled). Most typically, the distillation column is fluidly coupled to the absorber, receives a column feed stream and operates at a pressure that is at least 50 to 100 psi lower, and more preferably 100
25 psi to 300 psi lower than the operating pressure of the absorber. Seamless changeover of C3 recovery operation to C2 recovery operation (or *vice versa*) while maintaining 95% or higher C3 recovery for any level of C2 recovery is achieved by increasing the second reflux to the absorber while reducing the flow to the expander, with simultaneously diverting at least a portion of the absorber bottoms product flow to the distillation column.

30 One exemplary configuration for integrated NGL recovery with LNG liquefaction is depicted in **Figure 2** in which two columns operate at a pressure differential of about 300 psi,

and in which the absorber has an operating pressure about 600-700 psig. It should be pointed out that very high recovery of C2 plus components is possible in such plants and that the high pressure column operation is particularly beneficial in reducing the refrigeration duty in the LNG liquefaction plant. It should be further noted that such configurations and methods can be used for flexible C3 and C2 recovery. Exemplary compositions, temperatures, pressures, and flow rates of feed gas, product gas, and liquid product for typical operation are shown in the tables below. **Table 1** is a table depicting the overall mass balance for C2 recovery when the plant is operated on C2 recovery mode, while **Table 2** is a table depicting the overall mass balance for C3 recovery when the plant is operated on C3 recovery mode.

STREAM	FEED	LIQUID FROM NGL RECOVERY	RESIDUE GAS TO LIQUEFACTION
CO2	0.000	0.000	0.000
N2	4.569	0.000	5.007
C1	86.161	1.020	94.320
C2	5.046	51.017	0.641
C3	1.854	21.093	0.011
iC4	0.395	4.514	0.000
nC4	0.590	6.751	0.000
iC5	0.248	2.833	0.000
nC5	0.205	2.342	0.000
C6	0.224	2.565	0.000
C7	0.662	7.570	0.000
MMscfd	1,227	107	1,119
Barrel per Day		75,743	
Temperature, °F	120	115	-75
Pressure, psig	1,200	470	900

Table 1-C2 Recovery Mode

STREAM	FEED	LIQUID FROM NGL RECOVERY	RESIDUE GAS TO LIQUEFACTION
CO2	0.000	0.000	0.000
N2	4.569	0.000	4.768
C1	86.161	0.000	89.906
C2	5.046	0.867	5.228
C3	1.854	43.345	0.051
iC4	0.395	9.478	0.000
nC4	0.590	14.173	0.000
iC5	0.248	5.946	0.000

nC5	0.205	4.917	0.000
C6	0.224	5.384	0.000
C7	0.662	15.889	0.000
MMscfd	1,227	51	1,175
Barrel per Day		40,371	
Temperature, F	120	258	-60
Pressure, psig	1,200	440	886

Table 2- C3 Recovery Mode

In Figure 2, feed gas stream 1 enters the plant at about 1200 psig and 120 °F, and is cooled in heat exchanger 51 to typically -10 °F to -40 °F, forming stream 2 using letdown absorber bottoms stream 15, liquid stream 5 from separator 52, side reboiler stream 22 from the distillation column, and external refrigeration stream 32. It should be noted that any type of refrigeration system is suitable, including pure component cascade refrigeration cycles, mixed refrigerant cycles, or a combination of both systems. Further refrigeration is provided to the system with expansion using the turbo-expander 64 and various Joule-Thomson (JT) valves. The particularly high energy efficiency of contemplated processes and configurations is illustrated by the close temperature approaches of the heating and cooling curves in the minimization of loss work (or close temperature approaches) as shown in **Figure 3** and **Figure 4**. Here, the combined hot composite curve and the combined cold composite curve of the feed gas exchanger 51 and the reflux exchanger 54 are shown in Figure 3 and Figure 4 for C3 and C2 recovery, respectively.

The chilled feed gas stream 2 is separated in separator 52, forming a gaseous portion 3 and a liquid portion 4. The liquid portion 4 is letdown in pressure via JT valve 53 forming stream 5 typically at about -40 °F. During C3 recovery, stream 5 is heated in exchanger 51 to about 80 °F forming stream 6, using the heat content in the feed gas stream 1. Stream 6 enters the stripping section of the distillation column 61 for removal of the C2 and lighter components. The gaseous portion 3 from separator 52 is split into two portions. One portion (stream 7) is routed to the exchanger 54 to provide reflux 12 to the absorber via stream 9 and JT valve 55, while the other portion (stream 8) is expanded in turbo-expander 64 which generates power to operate compressor 65 to thereby form a chilled vapor stream 10, typically at -80 °F to -100 °F or lower. The chilled vapor is letdown in pressure to the absorber 58, which operates at 500 psig to 700 psig, typically at 600 psig.

For the different C2 recovery levels, the flow ratio (*i.e.*, ratio of stream 8 to stream 3), can be adjusted to maintain a high C3 recovery. **Table 3** shows examples of different flow ratios and the results of C3 and C2 recovery. Most notably, high C3 recoveries (typically over 98%) are maintained for all C2 recovery levels.

SPLIT RATIO (RATIO OF STREAM 8 TO STREAM 3)	C3 RECOVERY, %	C2 RECOVERY, %
0.7	98	85
0.8	98	62
0.9	99	31
1.0	99	25

Table 3

Absorber 58 is refluxed with two cold streams, with a first reflux stream 11 (top reflux) supplied by stream 27 from the distillation column 61, and a second reflux stream 12 from exchanger 54. Using twin reflux streams and suitable flow ratios, high C3 recovery can be maintained for the various levels of ethane recovery. During C3 recovery, the C2 content in the NGL product from the distillation column is lowered by increasing the column bottom temperature using heat supplied from the side reboiler and the bottom reboiler 63.

The absorber produces an overhead vapor stream 28 at about -100 °F to -110 °F and a bottoms stream 14 at about -90 °F to -100 °F. The overhead vapor is compressed by residue gas compressor 65 using power generated by turbo-expander 64 forming a discharge stream 29, typically at about 900 psig and -70 °F to -80°F. It should be especially appreciated that compression of a cold vapor is energy efficient as the achievable compression ratio across the compressor is significantly higher than that using a warm vapor of heretofore known plants. Thus, contemplated processes produce a high pressure and cryogenic temperature vapor that can be fed to the LNG liquefaction plant 67 for LNG production forming stream 30 at about -255°F to -260 °F.

The absorber bottoms stream 14 is letdown in pressure in JT valve 59 to about 460 psig, and is chilled to about -100 °F forming stream 15. During C3 recovery, this cold stream is used to provide a least a portion of the cooling duty of feed exchanger 51 and the reflux duty in condenser 62 to form streams 17 and 18, respectively. The overhead condenser 62 typically includes a heat exchange coil that is integral to the distillation column, generating internal reflux stream 19 to the rectification section of the distillation column. Alternatively,

the integral condenser 62 system can also be replaced by an external system, which would include an external heat exchanger, a separator and a reflux pump (not shown).

During C2 recovery, at least a portion of absorber bottom stream 14 is routed directly to the top of the distillation column for absorption of the C2 and heavier components in the distillation column. In this operation, JT valve 59 is partially, and more typically entirely closed and JT valve 60 is partially, and more typically entirely open forming stream 20, routing the cold absorber bottoms to the distillation column for C2 recovery. Thus, using various process streams within the NGL recovery plant, it should be recognized that the temperature profile of distillation column 61 can (gradually) vary between C3 recovery and C2 recovery as exemplified in **Table 4** below.

TEMPERATURE PROFILE OF 61	C2 RECOVERY	C3 RECOVERY
Top (Stream 24)	-100 to -120°F	-35 to -45°F
Bottom (Stream 25)	-90 to -120°F	240 to 270°F

Table 4

The NGL product composition is controlled with a side reboiler integral to exchanger 51, and a bottom reboiler 63 using an external heat source 34. The distillation column 61 produces an NGL bottoms products 25 (C2 plus and/or C3 plus) and an overhead vapor stream 24 that is compressed in compressor 66 to about 500 psig to 700 psig, or as needed to enter the absorber as a top reflux to form stream 26, further cooled in heat exchanger 54, and then used after pressure reduction in JT valve 56 as the first (top) reflux 11 to the absorber. Refrigerant stream 31 is supplied from an external refrigeration unit to the exchanger 54 for cooling and partially or totally condensing this recycle stream. Heated refrigerant stream 33 is later returned to the refrigeration unit.

It should be appreciated that recycling the overhead stream from the distillation permits recovery of the desirable NGL components allowing the distillation column to operate at the most efficient pressure for fractionation of separation of the desirable NGL components. Furthermore, and as illustrated in **Figure 5**, contemplated configurations take advantage of the relative volatilities of the NGL components (that is C1 to C2) in the columns using the pressure differential as compared to known configurations. Here, both the present configurations and the known configurations are compared for a column operating at 600 psig pressure for C2 recovery. The relative volatilities of the NGL components of the known configurations are represented by curve A, which drops to a very low value of 2 in the mid

section of the column. These low relative volatilities are the primary reason of lower separation efficiency and lower NGL recoveries, even with a large number of fractionation trays. In contrast, employing two columns operating at different pressures, with the first column (absorber) operating at a high pressure of 600 psig (curve B) and the second column
5 operating at 450 psig (curve C), dramatically increases relative volatilities of the NGL components (*e.g.*, to values of over 10), resulting in higher separation efficiency and higher NGL recoveries.

Thus, it should be recognized that the inventor discovered an efficient and flexible configuration and process that produces a high pressure cryogenic vapor stream (containing
10 predominately methane) suitable for feeding to an LNG liquefaction plant, and that further produces a liquid stream containing predominantly ethane and heavier hydrocarbons. Most typically, contemplated configurations and processes can achieve 95% or higher propane recovery when operating in a propane recovery mode, and can also achieve up to 50 to 85% ethane recovery when operating in an ethane recovery mode without substantial reduction
15 (*i.e.*, less than 5% absolute in reduction) in propane recovery. Contemplated configurations and processes can also provide a seamless and gradual changeover from a C3 recovery operation to a C2 recovery operation (or *vice versa*) by only adjusting flow ratios of streams to the absorber and the distillation column.

Viewed from a different perspective, the inventor discovered that high C2 and C3
20 recovery from a feed gas with relatively high pressure (*e.g.*, between about 800 psig to 1600 psig) can be realized by operating an absorber in a gas processing plant at a higher pressure than a distillation column (*e.g.*, a demethanizer or deethanizer), and in which a compressor is used to recycle the distillation column overhead to the absorber. In such configurations, the absorber bottoms product is preferably expanded to provide cooling for the feed gas and the
25 reflux stream. The overhead vapor from the absorber is then further compressed with power provided by the turboexpander, forming a feed gas to an LNG liquefaction plant without further recompression. Further components and considerations related to some aspects of the inventive subject matter are disclosed in our copending U.S. patent application with the serial number 10/478705, which is incorporated by reference herein. It is still further contemplated
30 that the configurations according to the inventive subject matter may find wide applicability in plants where high propane and ethane recovery are desirable, and where feed gas is available at a pressure greater than about 800 psig.

With respect to the feed gas, it is contemplated that numerous natural gas sources are suitable, including non-associated gas field production or gas fields associated with oil production, whether they are located on-shore or offshore. Consequently, the pressure of contemplated feed gas streams may also vary considerably, and where desirable, lower pressures may therefore be increased using boosters or compressors. However, it is generally preferred that appropriate feed gas pressures for plant configurations according to Figure 2 will generally be in the range between about 800 psig and about 1600 psig, and that at least a portion of the feed gas is expanded in a turboexpander to provide cooling and/or power for the residue gas recompression.

In especially preferred aspects, the absorber is configured to separately receive a first and a second portion of a feed gas vapor and a distillation column overhead, wherein the first portion of the feed gas vapor and the distillation column overhead provide the reflux to the absorber. In such configurations, a flow control adjusts the ratio of at least one of the first and second portions of the feed gas vapor to produce the desired recovery levels of ethane.

Among other advantages, it should be recognized that an optimum flow ratio of the first and second flow of the feed gas is employed for the variable C2 recovery while maintaining a high C3 (95% or above) recovery. Most typically, the flow control will include one or more manual and/or automatic valves, which are most preferably operated using microprocessor-based control equipment well known in the art. Such control units may operate entirely automatic without user intervention and may be configured to receive compositional information by sensors. Alternatively, or additionally, an operator may also provide compositional information based on a known or determined feed gas composition, and/or flow control may also be regulated at least in part by product volume flow in one or more products.

The absorber preferably produces an overhead vapor product that is predominantly methane (*e.g.*, at least 85%, more typically at least 90%, and most typically at least 93%) at cryogenic temperature (-80°F or lower), which is further compressed in a compressor using power generated by the turbo-expansion of the feed gas. Such configuration produces a high pressure cryogenic vapor at 800 psig to 900 psig or higher that can advantageously be directly fed to the LNG liquefaction plant. It should be recognized that compressing a cryogenic vapor from the high pressure absorber is more energy efficient, and therefore allows operation of an

LNG liquefaction plant with relatively low, and more typically no additional compression of the feed gas.

It should also be recognized that at least a portion of the absorber bottoms product is used as a lean oil for C2 absorption in the distillation column during C2 recovery. In such configurations, the flow to the feed cooler is reduced or stopped, thereby directing most or all of the absorber bottom product to the distillation column. C2 recovery in such configurations increases when the first portion of the feed gas vapor increases relative to the second portion of the feed gas vapor. Thus, preferred configurations permit a seamless changeover of the C3 recovery operation to the C2 recovery operation (or vice versa) while maintaining 95% or higher C3 recovery. The bottom product of the absorber is preferably expanded in a range of 50 psi to 350 psi, thereby chilled by Joule-Thomson effect to -90 °F to -130 °F. It is also contemplated that the cooled and expanded bottoms product stream is fed as the distillation column feed stream into the distillation column, and it is further contemplated that the expanded bottoms product stream may further provide cooling for the feed gas and reflux to the column, and at least of a portion of the expanded absorber bottoms product may also be routed directly to the distillation column for C2 absorption during the C2 recovery operation.

The distillation column in preferred configurations typically comprises a demethanizer or deethanizer column and produces an overhead stream, which is compressed, cooled, and fed to the absorber as the first absorber reflux stream. It should be noted that the compressor can be located in various locations between the absorber and distillation column, and that this reflux stream can be fed to various locations in the absorber for C3 recovery and/or C2 recovery. In yet further contemplated aspects, an external refrigeration unit is fluidly coupled to feed gas exchangers and reflux exchangers to supply feed gas chilling and column reflux duties. Such external refrigeration unit can be implemented using numerous configurations well known in the art using pure components or mixed refrigerants. For example, a cascade refrigeration process may employ heat exchange of the natural gas with several pure component refrigerants having successively lower boiling points. Alternatively, a single mixed refrigerant with multiple pure components can be used by evaporating the refrigerant at several different pressure levels. Also, where desirable, natural gas cooling can be achieved by expansion of the natural gas using either Joule-Thomson expansion or expansion turbine.

It is also contemplated that an external refrigeration unit can be employed to cool at least one of the first and second reflux streams, and may further cool at least one of the natural gas feed and a vapor portion of the natural gas feed. Where C2 recovery is particularly preferred, it is contemplated that the first lean reflux stream from the feed gas may be fed into the absorber as a subcooled liquid (*i.e.*, a liquid cooled below its bubble point temperature), and that the distillation column is a demethanizer.

Consequently, in one aspect of the inventive subject matter, a method of operating a plant includes a step of providing an absorber and a distillation column, wherein the absorber receives a plurality of absorber feed streams and provides a bottom product to the distillation column, and an overhead absorber vapor that is used as feed to the LNG liquefaction. In another step, at least one of the feed streams is split into a first and second portion, wherein the first and second portions are introduced into the absorber at different locations, and in still another step, the flow ratio between the first and second portions is used to control the degree of recovery of C2 components in the bottom product of the distillation column. In yet a further step, the absorber bottoms product is used as a lean oil for absorption of the C2 component in the distillation column.

Viewed from a different perspective, contemplated configurations may also be used in a method of increasing throughput and/or energy consumption in an LNG liquefaction plant that is coupled to a natural gas recovery plant having an absorber and a distillation column. Such methods will typically include one step in which a first reflux stream is provided to the absorber, wherein the first reflux stream comprises an overhead product from the distillation column. In another step, a bypass is provided upstream of a turbo expander, wherein the bypass receives a vapor portion of a cooled natural gas and provides the vapor portion to the absorber, and in yet another step, the pressure of the vapor portion is cooled, at least partially condensed and reduced in pressure before the vapor portion enters the absorber as a second reflux stream. In a still further step, a heat exchanger cools at least one of the first and second reflux streams using an external refrigerant. Therefore, a method of operating a plant may include one step in which an absorber and a distillation column are provided. In a further step, a cooled lean overhead product from the distillation column is fed to the absorber as a first reflux stream, and in another step, the pressure of a cooled vapor portion of a natural gas feed is reduced via a device other than a turbo expander, wherein the cooled vapor portion that is

at least partially condensed or subcooled, reduced in pressure and fed to the absorber as a second reflux stream.

With respect to the remaining components of contemplated configurations (*e.g.*, heat exchangers, pumps, pipes, valves, compressors, expanders, etc.) it should be appreciated that such components are well known to the artisan and that all known/commercially available components are deemed suitable for use in conjunction with the teachings presented herein. Furthermore, the term "about" where used in conjunction with a numeral refers to a numeric range that encompasses +/- 10% of the numeral, inclusive. For example, the term about 10% refers to a range of 9% to 11%, inclusive.

Thus, specific embodiments and applications for high propane and ethane recovery processes and configurations have been disclosed. It should be apparent, however, to those skilled in the art that many more modifications besides those already described are possible without departing from the inventive concepts herein. The inventive subject matter, therefore, is not to be restricted except in the spirit of the appended claims. Moreover, in interpreting both the specification and the claims, all terms should be interpreted in the broadest possible manner consistent with the context. In particular, the terms "comprises" and "comprising" should be interpreted as referring to elements, components, or steps in a non-exclusive manner, indicating that the referenced elements, components, or steps may be present, or utilized, or combined with other elements, components, or steps that are not expressly referenced. Furthermore, where a definition or use of a term in a reference, which is incorporated by reference herein is inconsistent or contrary to the definition of that term provided herein, the definition of that term provided herein applies and the definition of that term in the reference does not apply.